Optimization and performance improvement of Lekhwair natural gas sweetening plant using Aspen HYSYS

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Abstract

Amine gas sweetening process is widely used to remove acid gases (H_2S and CO_2) from natural gas. However, this process is energy intensive due to the high requirements of heating and cooling. Moreover, reducing its energy requirements by optimizing its operating parameters is limited. Therefore, process modifications can make a considerable reduction in energy consumption and hence reducing the operating costs. This research reviews the current operation of an existing plant (Lekhwair plant, Oman) considering the main operating parameters (lean amine circulation flow rate, temperature and concentration) and proposes some modifications to the existing plant to increase its profitability and sustainability. The simulation and sensitivity analysis are carried out using Aspen HYSYS v7.3. The operating capacities of some equipment are reviewed to assess the possibility of changing the operating parameters along with investigating the occurrence of common operational problems like foaming. Two modifications (conventional split-loop and modified split-loop) are simulated and discussed. A comparison between them and the current process are carried out in terms of profitability and sustainability. The conventional split-loop has been found to save up to 50% of the current operating expenses with only around £175,000 increase in capital investment, and a penalty of 1.0 ppm of H_2S concentration in the sweet gas, which is still well below the pipeline gas specification. Finally, a sulphur recovery process is proposed to make the plant more sustainable and environmentally friendly along with proposing two modifications to the conventional sulphur recovery process. Even though the conventional and the proposed modified sulphur recovery processes are not economically profitable, the modified sulphur recovery process is more sustainable as its carbon footprint is lower than the conventional process.

Keywords: Natural Gas; Gas Sweetening; Aspen HYSYS; Process Optimization; Profitability Analysis; Sulphur Recovery.

1. Introduction

Natural gas is a prime source of energy, which is widely used as an industrial and domestic fuel. To make natural gas suitable and environmentally safe to use, it is crucial to purify it from all contaminants that can affect its utilization and optimal energy capacity. These contaminants can also cause problems such as corrosion, freezing, plugging, erosion, health and environmental hazards (Ghanbarabadi and Khoshandam, 2015). The acid gases of hydrogen sulphide (H_2S) and carbon dioxide (CO_2) are impurities existing in natural gas brought up from wellhead (oil wells with associated gas or gas wells). The gas is considered sour if its H_2S content exceeds 5.7 mg m^{-3} (3.8 ppm volume-based). The process for removing H_2S and CO_2 from a natural gas stream is referred to as gas sweetening (Abkhiz et al., 2014). With the increasingly strict environmental regulations on emissions from natural gas treatment plants and also the market demand for high quality natural gas, the gas sweetening process has become mandatory (Rezakazemi et al., 2011). The most widely used method for acid gases removal is amine gas sweetening process with more than 50% of the current acid gas removal technologies use aqueous solutions of alkanolamines. However, this gas sweetening process is energy intensive especially for amine regeneration (Wang et al., 2015). Therefore, optimizing the amine gas sweetening process could result in great energy savings and thus remarkable economic benefits for the existing gas sweetening plants. Depending on their molecular structure, the amines are characterized into three main groups: primary, secondary, and tertiary. Mixtures of amines are also used in industry. However, tertiary

amines, such as methyldiethanolamine (MDEA), have several advantages over primary and secondary amines. These advantages include lower vapour pressure, lower heats of reaction, higher resistance to degradation, and less corrosion problems. Another important advantage is absorption selectivity of H_2S preferentially to CO_2 (Qiu et al., 2014). Therefore, MDEA will be used in this work.

The waste acid gases from the process are normally incinerated or flared to the environment. In the incinerator, H_2S is converted to SO_2. The flaring of sour gases not only means that natural gas resources are wasted, but also results in the release of pollutants into the atmosphere. CO_2 is a major contributor to global warming and SO_2 causes acid rain; accelerated corrosion of buildings and reduced visibility (Abu-Eishah et al., 2013). Recently, most of the publications on the pre- or post-treatment of natural gas is concentrating on CO_2 capture either using physical or chemical solvents (Budzianowski and Koziol (2005), Budzianowski (2011a), Budzianowski (2011b), Budzianowski (2015)). However, the major concern of industry is always safety and hence the removal of H_2S, which is highly toxic, is still constraining the design of natural gas production and processing facilities (Duissenov, 2013). Therefore, optimizing the performance of existing processes or investigating possible process modifications should always take into consideration keeping H_2S concentration in treated natural gas under the allowable discharge concentration.

Lekhwair amine gas sweetening plant is built and operated by Petroleum Development Oman (PDO) Company in Lekhwair field (Northwest of Oman) to treat the produced natural gas in the field, which is rich in H_2S and CO_2. The oil and water are separated and the gas goes to a pre-treatment process before flowing to the amine gas sweetening plant. This plant was commissioned in 2012 to remove H_2S from the produced natural gas and is designed to reduce the H_2S concentration of the sour gas from 500 ppm to the maximum allowable discharge concentration of 5 ppm (volume-based) with an inlet gas flow rate of 5.0 million standard cubic meter per day (MMSCMD) (PDO Company, 2012). The current feed flow rate of acid gas is less than 0.5 MMSCMD and the the H_2S concentration in the feed gas is 200 ppm. Due to operating at lower flow rate, the concentration of H_2S in the treated sweet gas is around 0.04 ppm, which is acceptable but considered low. However, this very low concentration of H_2S in sweet gas causes the concentration of SO_2 in the flare of waste gases to be always above the target. Also, there is a chance for an increase in acid gases due to new wells being drilled and reservoir change (Shihab et al., 2013). Additionally, as the plant is considered new, there is a requirement to review the current plant operation and investigate any possible optimisation and modification that can save energy and lead to reduction in the flared waste gases without violating the sweet gas specifications.

Reduction of the operating costs and increasing the environmental sustainability of the process are two critical concerns in any oil and gas processing facility. The reduction of the energy consumption by optimizing the operating parameters is proved to be limited. However, process modifications can result in a considerable reduction in the energy requirement of the sweetening process. There are several approaches to reduce energy consumption such as, replacing amine type, modifying column internal or modifying process flowsheet. In a conventional amine gas sweetening process, the amine flows as a single stream from the bottom of the absorber to the top of the stripper, and from the bottom of the stripper to the top of the absorber. There are however possibilities to have multiple feeds or draws in both the absorber and stripper. Such configurations are called split-flow or split-loop configurations. The main advantage with a split-

loop configuration is reducing heat consumption in the stripper. This reduction is because that only a portion of the amine needs to be fully regenerated. Another explanation is that the driving force especially in the absorber is reduced (Sabbagh et al., 2014).

Most of the recent publications on split-loop configuration are interested in improving CO_2 removal from natural gas. A survey of amine gas sweetening process modifications for CO_2 removal is reported by Cousins et al. (2011). Moreover, very few calculations of CO_2 removal from natural gas based on split-loop configuration have been published (Sabbagh et al., 2014). Vozniuk (2010) has simulated the traditional amine gas sweetening process with and without split-loop configuration using Aspen HYSYS v7.0 by applying amine package and Kent-Eisenberg model. It was one of the first reports that quantitatively investigated the effect of split-loop configuration. However, up to the authors' knowledge, there are almost no recent publications investigating the effect of these modifications on H_2S removal.

So far, most of the published studies considered only the qualitative evaluation of the proposed process modifications. Therefore, this study is pursuing to quantitatively examine the reduction of energy consumption in an existing amine gas sweetening plant using Aspen HYSYS as an optimization tool. Part of the novelty of this research relies on using a commercial process simulator (Aspen HYSYS v 7.3), where different fluid packages (amine and DBR amine) are applied and evaluated against the actual process to choose the most suitable package. This approach takes advantage of commercial process simulators (specially tailored fluid packages, reliability, and robustness) and the flexibility of creating and modifying process flow diagrams for the modelling and simulation purposes. The optimization procedure proposed is successfully applied to the amine gas sweetening process of Lekhwair plant, where different alternatives are available for the streams, equipment and process conditions. Another part of this research novelty is proposing the modified split-loop configuration with heat integration, which has not been simulated before. Also, different modifications of conventional sulphur recovery process are considered to ensure a stable and reliable sulphur production process.

This paper firstly considers the optimization of Lekhwair gas sweetening process to reduce the operating expenses, taking into consideration equipment limitations. Secondly, different process modifications (conventional and modified split-loop configurations) are simulated, optimized. Finally, reducing the waste gas flare by converting H_2S into a valuable product (e.g. sulphur) taking into consideration the conditions of the waste gas and the required modifications in the process. A comparison between the proposed different modifications is carried out in terms of operating costs, capital costs and carbon footprint.

2. Methodology

Aspen HYSYS v7.3 is used to study, simulate and optimize Lekhwair amine gas sweetening plant. In this study, DBR amine package is used rather than traditional amine package. The improved DBR amine package is a new package added to HYSYS v7.3 and is similar to the amine package, but is independently coded and can be updated anytime. It is also better handles the reaction models of MDEA and CO_2 mixtures (AspenTech, 2011). The selected thermodynamic model is the modified Kent and Eisenberg model, which is used to calculate the vapour pressures of H_2S and CO_2 above amine solutions. This model assumes non-ideal liquid phase with the activity factors being implied in equilibrium constants. This model is also useful for molar loadings greater than one (Boroojerdi et al., 2013). Patil et al. (2006) have studied in details the accuracy of predicting CO_2 and H_2S solubility in aqueous MDEA solutions using the

modified Kent and Eisenberg model and the predictions were validated against experimental data and showed to be in very good agreement. The performance of the existing PDO Lekhwair amine gas sweetening plant is firstly evaluated using both packages. The comparison between the actual plant data and simulation data for the main operating parameters showed a reasonably good agreement. However, HYSYS is estimating higher H_2S concentration in the sweet gas as can be seen in Table 1. The amine package has not fully converged because the sweet gas loading (acid gas moles/MDEA moles) is greater than one. For amine package, the limitation for sweet gas loading is less than one. However, DBR package skips this limitation and hence, it is selected. Further evaluation of the two packages in terms of the flow rates of different components and the duties of different utilities will be given in the results and discussion section. The simulation and optimization of PDO Lekhwair amine gas sweetening plant is divided into two main parts. In the first part, the current operation is reviewed and the important operating parameters (MDEA circulation rate, concentration and inlet lean amine temperature) are optimized taking into consideration the current equipment limitations. The second part includes retrofitting and process modifications that need capital investments.

Table 1

Comparison between HYSYS results and plant data

	Plant data	HYSYS Amine package	HYSYS DBR amine package
H_2S conc. in sweet gas (ppm)	0.040	0.200	0.074
CO_2 conc. in sweet gas (vol %)	--*	0.159	0.059

*CO_2 analyzer not yet commissioned (PDO, 2012)

The existing gas sweetening process is considered as the base case (Case 1) and has been simulated using Aspen HYSYS as shown in Figure 1. This is an MDEA absorption process consisting of two parallel trains, each with a capacity of 2.5 million standard cubic meter per day (MMSCMD). A 41 wt. % MDEA solution is circulated at a flow rate of 18.75 $m^3\ hr^{-1}$ to meet the gas outlet specification of maximum 5.0 ppm of H_2S. Both trains are identical in design and capacity. The raw natural gas to each train enters a filter coalescer to remove any liquid droplets and carryover from the upstream conditioning processes in order to minimize liquid hydrocarbon that can condense and cause foaming in the system. The raw gas then enters the bottom of the gas absorption column (25 actual stages) at a temperature of 51°C and a pressure of around 7154 kPa, and flows upwards counter currently contacting lean MDEA solution that enters from the top of the column at a temperature of 57°C.

The inlet sour gas to the absorber consists mainly of methane (80.1 mol %), ethane (11.4 mol %), propane (3.25 mol %), H_2S (200 ppm) and CO_2 (2600 ppm). The warm MDEA solution and the exothermic reaction cause the sweet gas to warm before leaving the top of the absorber. Rich MDEA solution, saturated with dissolved hydrocarbons, flows to a flash tank which operates at a pressure of around 551 kPa gauge. The rich MDEA from the flash tank gets heated to approximately 119°C by exchanging heat with the lean MDEA stream coming from the bottom of the stripper (17 actual stages) at a temperature of 124.8°C through the lean/rich heat exchanger. Then, the heated rich MDEA solution flows to the stripper as a feed. The stripper operates at a bottom temperature of approximately 124°C and a pressure of 206.3 kPa; and at top temperature of approximately 119°C and a pressure of 196.3 kPa. The stripped gas from the rich

MDEA stream leaves the top of the stripper and is partially condensed and subcooled to approximately 40°C in the stripper overhead condenser. The condensed liquid is returned to the stripper as reflux. The acid gas leaving the top of the reflux accumulator is sent to the acid gas incinerator to get burnt and converted to SO_2 before venting. The bottom of the stripper, which is lean MDEA solution, is removed by the lean solvent booster pump. The pump circulates the lean amine through the lean/rich heat exchanger, the air cooler, and provides sufficient suction pressure for the operation of the lean solvent charge pump. A water makeup is added to the lean MDEA to adjust its concentration. The lean MDEA is cooled in the air cooler to approximately 57°C, and pumped to the absorber as a feed. The process main parameters (e.g. MDEA solvent circulation rate, concentration, and the inlet lean amine temperature to the absorber) will be reviewed and optimized.

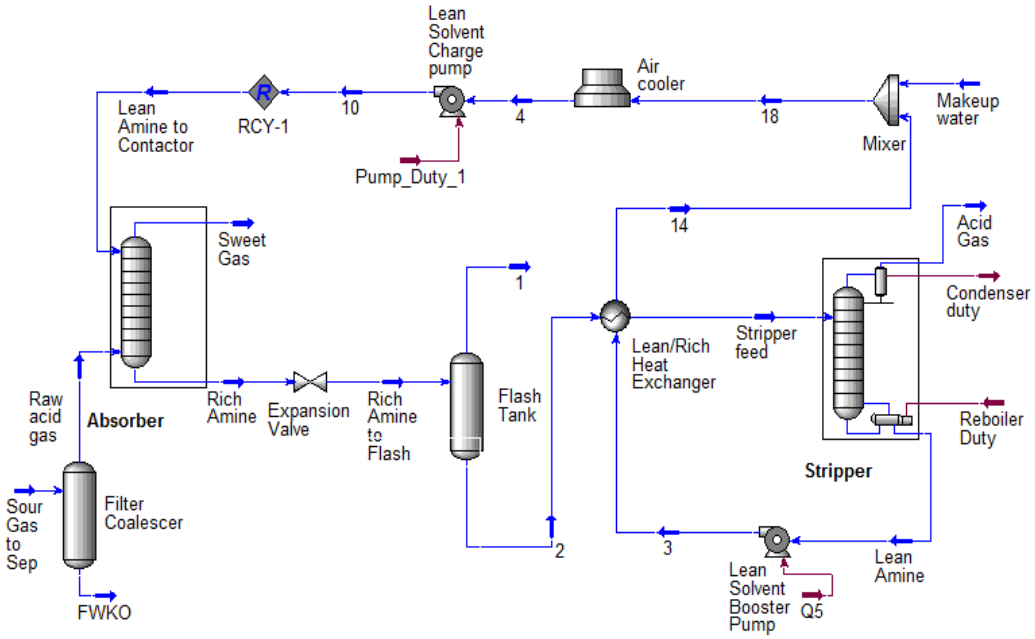

Fig. 1. Aspen HYSYS simulation of PDO Lekhwair amine gas sweetening plan.

2.1. Modifications to the existing process

To get lower outlet concentrations, more energy needs to be supplied, and hence, more operating costs. To overcome the increase in energy consumption, two process modifications are proposed to meet the required gas purity at lower energy costs. These two process modifications are compared in terms of the outlet H_2S concentration and the regeneration duties. For retrofit and comparison purpose, the two modifications will be made to the base case (Case 1) without any changes to the recommended parameters after process review.

2.1.1.Side draw from the stripper to the absorber (conventional split-loop)

This side draw modification (named Case 2) is simulated using Aspen HYSYS (Fig. 2) and an optimization procedure is followed to lower energy consumption. The optimization procedure is summarized in a simple algorithm shown in Figure 3. The first variable that needs to be optimized is the side draw flow rate. This variable is converted to a fraction of the total gas inlet to the absorber, and will be called side fraction. At selected circulation flow rate and amine concentration, the side fraction is increased. Increasing the side fraction, and hence the solvent

flow rate drawn from the stripper column can reduce the reboiler heating requirement due to the low solvent flow rate passing the reboiler. On the other hand, the absorption performance is expected to decline due to the low lean solvent flow rate entering the top section of the absorber. Moreover, the side draw is semi lean amine that has acid gases unstripped in the regeneration column. Therefore, H_2S and CO_2 concentrations in the sweet gas are expected to increase. In this case, as the last step in the algorithm, the reboiler duty needs to be adjusted to meet the outlet specifications. The final heating duty after adjustment is expected to be still less than that required in the base case. However, if the outlet concentration of H_2S after side draw optimisation is well below the maximum specification allowed (5 ppm), the reboiler duty can remain unchanged.

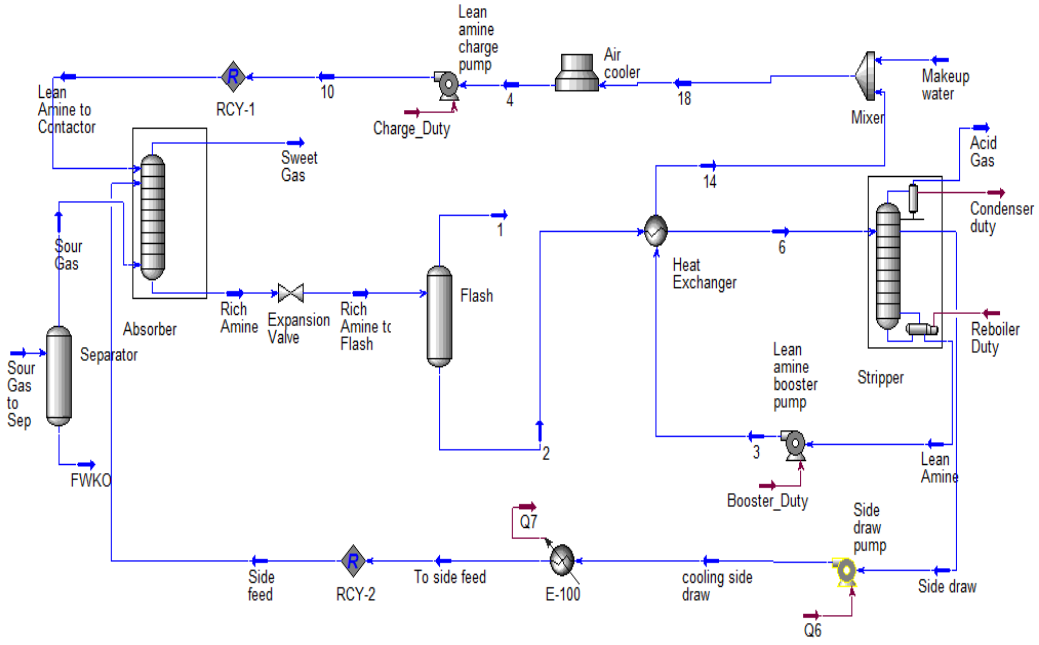

Fig. 2. Aspen HYSYS simulation of the conventional split-loop process modification (Case 2).

The optimum location of the side draw in the stripper and side feed in the absorption column can be optimized by HYSYS by carrying out different sensitivity analysis. For any optimisation, good *initial estimates* of the parameters are required. In this work, the main parameters are the side fraction, the intermediate side draw in the stripper, and the feed stage in the absorber. If other parameters are not available like solvent circulation rates or temperature, they can be estimated initially from heuristics from the literature or industrial experiences. The solvent circulation flow rate can be estimated by using Eq. 1 (Kohl and Nielsen, 1997) assuming acid gas loading is 80% of the inlet acid gas partial pressure:

$$L = (y_{in}\ G)/(0.8x_e) \tag{1}$$

Where, L = molar circulation flow rate (kmol h^{-1}); G = molar gas feed rate (kmol h^{-1}); y_{in} = inlet acid gas mole fraction; x_e = acid gas mole fraction in equilibrium with y_{in}. The initial estimates of the parameters for the optimisation are listed in Table 2.

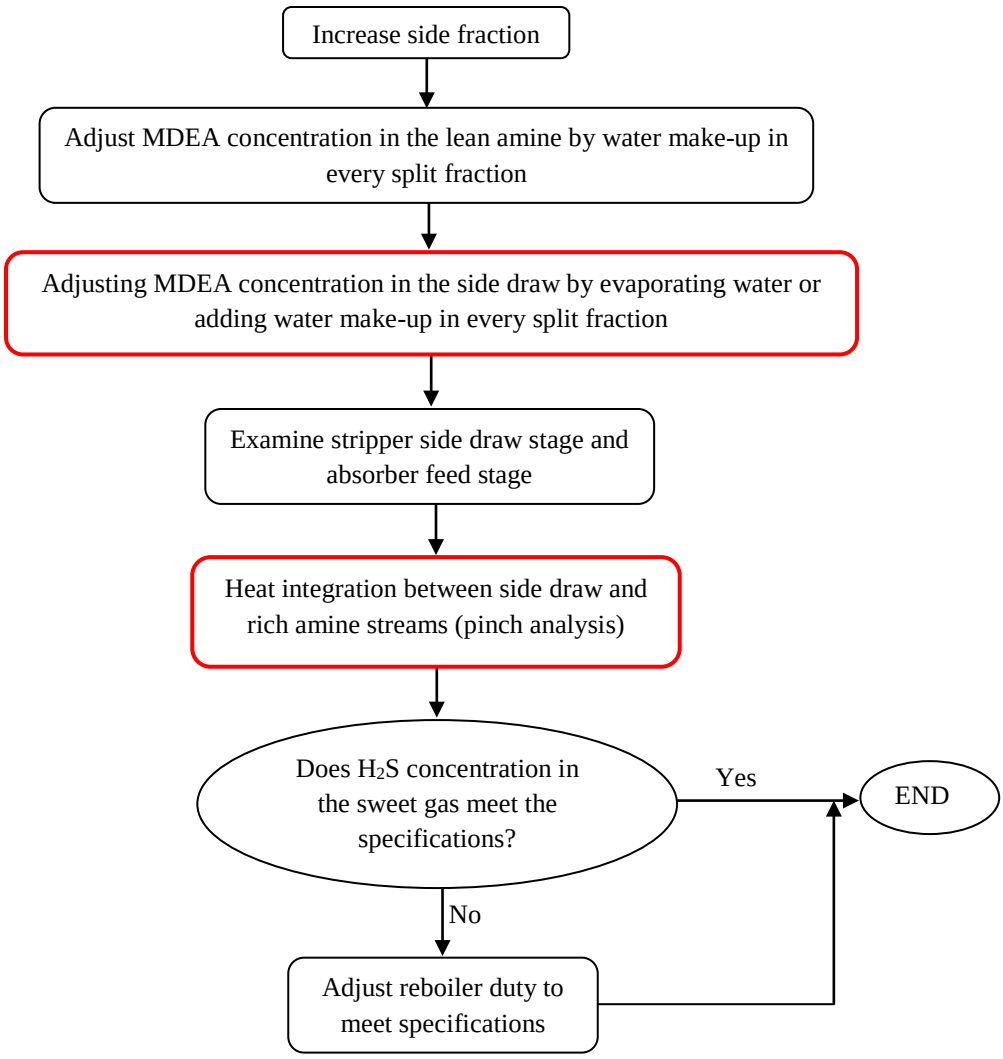

Fig. 3. Algorithm of side stream optimization (*Note: Steps highlighted in RED are extra steps for the modified side draw process (Case 3)).*

Table 2

Initial estimates of the parameters for the optimization.

Parameters	Initial estimates
Side fraction	0.1
Stripper intermediate side draw stage*	9
Absorber intermediate side feed stage*	13
Absorber lean amine inlet stage (fixed)	25
Absorber feed gas inlet stage (fixed)	1
Stripper feed gas stage inlet stage (fixed)	17
Reboiler duty (kW)	2170

* Stage numbering in both columns from bottom up.

This modification can make a considerable reduction in the regeneration duties. However, it was found that the MDEA concentration in the side draw is lower than that in the lean MDEA solution, and hence these two streams enter the absorber at different concentrations. Although it was found that increasing or decreasing MDEA amine concentration around 40 wt. % does not enhance the H_2S removal efficiency, the concentration of 41 wt. % has to be maintained in the system. Therefore, a modified split-loop needs to be proposed and examined. This additional process modification will be named Case 3.

2.1.2.Modified split-loop (Case 3)

The MDEA concentration in the side stream can be controlled by evaporating water until achieving 41 wt. % MDEA. The vaporization process can be simulated in HYSYS by using a heater and a separator and the overall flow diagram of this modified process is shown in Figure 4. A trial and error method is used to find the optimal heater outlet temperature. If the concentration of MDEA is above 41 wt. %, it can be diluted by adding make-up water. Another modification is being proposed to this process which is a heat integration opportunity that can reduce the operating cost of the cooler in the side draw stream. The side draw from the stripper is hot (ranging from 120 to 125°C at the current conditions), while the rich MDEA solution is relatively cold (around 51°C). The rich MDEA stream is currently exchanging heat with the lean MDEA stream coming from the stripper. Therefore, the stream can be split to provide process cooling to both streams (lean amine and side draw). The optimization algorithm (Fig. 3) is modified by introducing two more steps highlighted in red; adjusting the MDEA concentration by adjusting the heater outlet temperature, and heat integration possibility using pinch analysis.

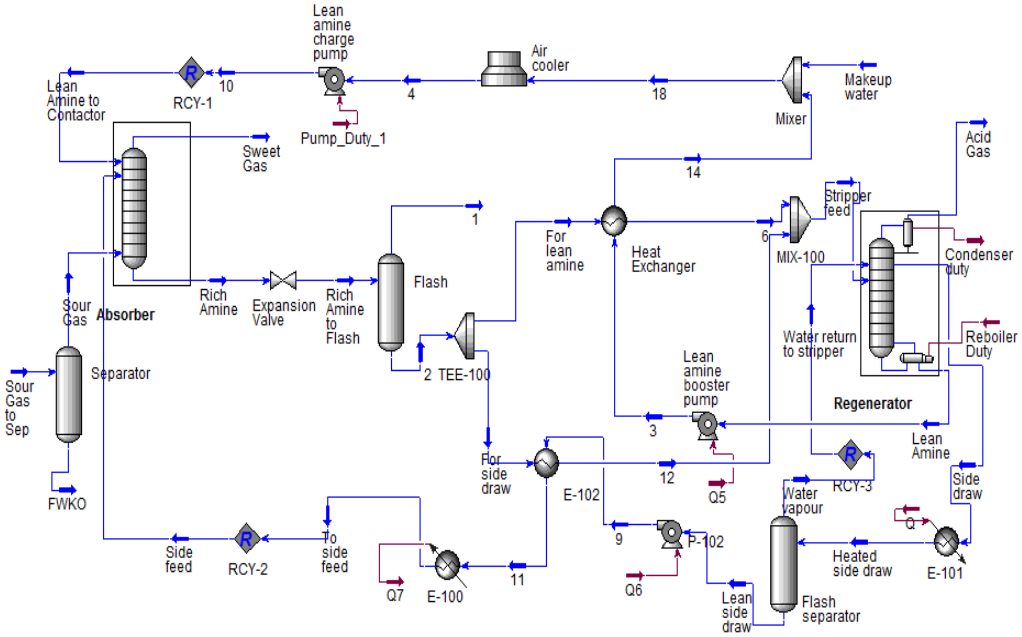

Fig. 4. Aspen HYSYS simulation of the modified split-loop process (Case 3)

2.1.3.Sulphur Recovery

Currently, H_2S from the stripper is sent directly to the acid gas incinerator and burnt and converted to SO_2. Due to the severe effects of SO_2 to human health and environment, sulphur recovery process is proposed to reduce the SO_2 emissions. This process needs to be constructed

and operated upstream of the incinerator. The remaining unconverted H_2S and SO_2 gases will be routed to the incinerator for final burning. Aspen HYSYS v7.3 is used for simulating the sulphur recovery process, as shown in Figure 5, and the required modifications using Ideal Aspen Properties package. The conventional sulphur recovery process consists of a furnace and waste heat boiler, reactors, heaters and condensers. As the acid gas is considered lean ($H_2S \leq 10$ mol %), the modifications (bypass the furnace or oxygen enrichment) are examined. The selection of the best case will be based on the best operating conditions as well as the project profitability that is discussed in the next section.

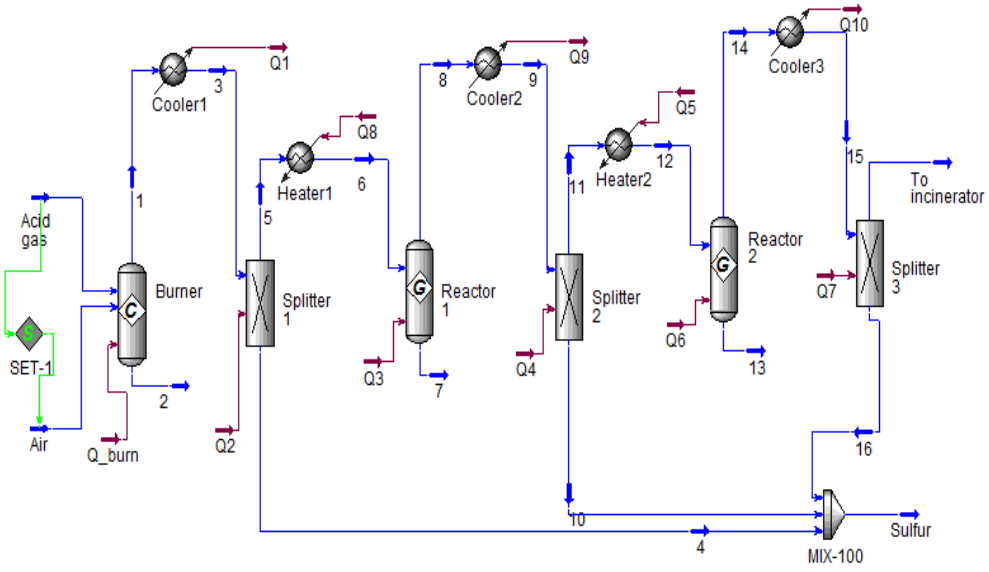

Fig. 5. Aspen HYSYS simulation of a conventional sulphur recovery process (Case 4)

2.2. Profitability Analysis

The total capital investment is the sum of the fixed capital investment and the working capital. The fixed capital investment of a chemical process plant includes the Inside Battery Limits (ISBL) which is the purchase cost of the major equipment items required for the process, and the other installation costs being estimated as factors of the equipment cost, the Outside Battery Limit (OSBL) cost which will be assumed as 30% of ISBL. For a preliminary design, where only the main equipment are considered, contingency and engineering charges will be assumed to be 25% of (ISBL + OSBL) costs. The fixed capital investment is the sum of ISBL, OSBL, engineering and contingency. The working capital is the capital required to operate the plant. A charge of 15% of the total capital investment will be considered for the working capital cost (Sinnott and Tawler, 2009). Start-up cost of 10% of the fixed capital cost will be included in the investment in the sulphur recovery process.

The accuracy of this type of estimate depends on the stage of the design and on the reliability of the data available on equipment costs. There are many methods that can be used for cost estimation; among the available methods, the factorial method is a relatively accurate for capital cost estimation (Sinnott and Tawler, 2009) and it will be followed in this project. Details of the factorial method is in Appendix A.

As the sweetening plant is located in Oman, there are differences in currency exchange, costs of shipping, local fabrication and construction. To capture these differences in the cost estimation, the estimated capital cost will be multiplied by a Location Factor. The location factor for Middle East is 1.07 (Sinnott and Tawler, 2009) and this factor is assumed to be the same for the year 2014.

The operating costs in this project include mainly utilities (fuels, electricity, steam, and cooling water) which are the most significant variable operating costs, catalysts for sulphur recovery process, and chemicals (MDEA amine and antifoam chemical). More details on estimating the operating costs are given in Appendix A. The annualized operating and capital costs will be used to compare the four cases.

The economic analysis of the proposed sulphur recovery process is evaluated using two simple measures of economic performance; Payback Period and Net Present Value (NPV) (Sinnott and Tawler, 2009). For Payback period, the shorter the period is, the more attractive the project is. One simple method of estimating the payback period is by dividing the total initial capital investment (fixed capital plus working capital) by the average annual cash flow:

$$Payback\ period = \frac{total\ initial\ capital\ investment}{average\ annual\ cash\ flow} \tag{2}$$

Net Present Value (NPV) of a project is the sum of the present values of the future cash flows:

$$NPV = \sum_{n=1}^{n=t} \frac{CF_n}{(1+i)^n} \tag{3}$$

Where CF_n = Cash flow in year n; t = project life in years; and i = interest rate.

2.3. Sustainability and carbon footprint calculation

Life Cycle Assessment (LCA) is an environmental management technique that helps translate life cycle thinking into a quantitative measure of environmental sustainability of processes or activities from raw material extraction through materials processing, manufacture, distribution, use, maintenance, and disposal or recycling on a life cycle basis (Azapagic and Perdan, 2011). In this study, as the raw materials, distribution, transportation and use stages are almost the same for all simulated processes; the processing stage is only considered for sustainability and carbon footprint calculation. The carbon footprint is calculated based on the actual CO_2 emission loads from Lekhwair production process (PDO, 2012). The carbon footprint is measured in kg of CO_2 equivalent ($kgCO_2e$) based on a function unit (e.g. 1.0 kg of sulphur, 1.0 ppm of H_2S in the sweet gas).

3. Results and Discussion

3.1. Plant data validation

Before optimizing the actual gas sweetening process, the simulation results for the sweet gas stream specifications and required duties using the two HYSYS packages are compared to the actual Lekhwair plant data. The comparison is summarized in Table 3. The comparison shows

that the two simulation packages predict the sweet gas stream specifications and the required duties very well. The magnitude of absolute error between actual and simulated data is less than 5%. Also, DBR amine package gives generally more accurate results.

Table 3

Simulation results for the sweet gas stream specifications using HYSYS Amine and DBR amine packages compared to the actual plant data

	Sweet gas stream specification				
	Plant data	Simulation results			
	Molar flow (kmol h^{-1})	Amine package (kmol h^{-1})	Error (%)	DBR amine package (kmol h^{-1})	Error (%)
Methane	2852.21	2830.12	0.77	2833.22	0.67
Ethane	387.19	400.40	3.41	401.25	3.63
Propane	112.50	114.67	1.93	114.30	1.60
n-Butane	47.04	48.66	3.45	48.54	3.19
i-Butane	40.60	40.07	1.30	40.10	1.23
i-Pentane	15.33	15.85	3.43	15.83	3.26
n-Pentane	9.10	9.50	4.39	9.50	4.40
Nitrogen	31.92	32.89	3.04	32.94	3.20
Total (dry basis)	3497.72				
Temperature (°C)	55.0	55.4	0.73	54.7	0.55
Pressure (kPa)	7140	7140	0.00	7140	0.00

	Required Duties (kW)				
		Simulation results			
	Plant data	Amine package	Error (%)	DBR amine package	Error (%)
Lean/rich heat exchanger	1350	1344	0.52	1351	0.07
Stripper condenser	1960	1971	0.56	1971	0.56
Stripper reboiler	2170	2170	0.00	2170	0.00
Lean solvent charge pump	125	127.3	1.80	126.4	1.12

3.2. Actual process optimization

The amine circulation flow rate is one of the most important parameters in the natural gas sweetening process. It has to be adjusted to meet the required sweet gas specifications and the system capacity. This flow rate is adjusted by a lean amine charge pump. This pump is designed to operate at a total depth head (TDH) of 750 m and 150 kW of power consumption. Also, its optimum efficiency ranges from 55 to 70%. The current flow rate of the solvent in the plant is 18.75 m^3 hr^{-1}, which causes the lean amine charge pump to operate outside this optimum range. The efficiency of the pump running at the current conditions is only around 30% with Total Depth Head (TDH) of 750 m and 120 kW of power consumption. The effect of operating away from the optimal efficiency range will speed up the wear and tear on the pump thereby reducing reliability. At 30% efficiency, which is 65% of the lower limit of optimal pump efficiency range,

will result in 50% lifetime reduction of seals, 20% lifetime reduction of bearings, 25% lifetime reduction of casing and impeller and approximately 100% increase of maintenance cost (Lou et al., 2012).

The amine circulation flow rate has to be increased to shift the pump performance to the optimal range to reduce potential damages to the pump. However, there could be some penalties in the process operation due to the increase in the circulation rate. A sensitivity analysis is carried out using HYSYS (Fig. B1, Appendix B) to investigate the effect of the amine circulation rate on the concentration of acid gases (H_2S and CO_2) in the sweet gas as well as the regeneration duty requirements, keeping the number of trays in the columns and other parameters fixed. The sensitivity analysis showed that as the solvent circulation flow rate increased, the H_2S and CO_2 concentrations in the sweet gas decreased. It was also observed that below amine circulation rate of 58.75 $m^3\ h^{-1}$, the overhead of the acid gases rise rapidly, and this can be a good condition for processes requiring CO_2 rejection in the sweet gas. For a circulation rate of 80 $m^3\ h^{-1}$ and above, the H_2S concentration in the sweet gas tends to become unchanged.

In addition to this, the acid gas loading, which is defined as the number of moles of acid gas absorbed per mole of amine (Kazemi et al., 2014), can play a role in controlling the solvent circulation rate in the system. By carrying out a sensitivity analysis using HYSYS (Fig. B2), it was found that increasing the circulation rate decreased the H_2S and CO_2 loading in the rich amine. So, low circulation rate will increase the possibility of insufficient H_2S and CO_2 removal from the raw gas and the amine becomes more corrosive. On the other hand, it was found that the hydrocarbons in the rich amine increased with increasing amine circulation rates. These hydrocarbons, mainly CH_4, can cause foaming in the regeneration column and an upset in the downstream process (Abdulrahman and Sebastine, 2013). Furthermore, regeneration reboiler duty increased due to increased flow rate entering the regeneration column (Fig. B3).

Increasing the amine circulation rate resulted in shifting the pump performance curve to its optimum operation and hence will increase its lifetime. It also enhanced the acid gases removal efficiency in the absorption column. On the other hand, it increased the regeneration heating and cooling requirements to almost double, increased the electric motor drive power requirement for the pump, and increased the possibility of having more foaming upsets. Therefore, a trade-off between these factors needs to be carried out, and probably more modifications in the system and changes in the operating parameters need to be applied to optimize the operation. The optimum circulation rate can be selected as 40 $m^3\ hr^{-1}$, which is just at the best operating range of the pump. At this flow rate, the increase in the operating costs can be summarized in Table 4.

Table 4

Comparison between the base case and the case of circulation rate of 40 $m^3\ hr^{-1}$.

	Base case (18.75 $m^3\ hr^{-1}$) (£ yr^{-1})	Circulation rate at 40 $m^3\ hr^{-1}$ (£ yr^{-1})
Electricity requirements	31,412	37,695
Heating requirements	197,600	410,681
Cooling requirements	4,920	10,457
Total	233,932	458,833

As can be seen from the Table 4, the operating expenses increased by around £200,000 per year when the circulation rate is increased to 40 m^3 hr^{-1}. However, the installed capital cost of the pump as per the design economics (PDO Company, 2012) is around £608,000. Normally, the annual cost of a pump maintenance accounts about to 3% of the installed capital cost of the pump (Sinnott and Towler, 2009). Therefore, when the pump is operating outside the optimal efficiency range, this cost will increase to 6%. Moreover, the cost of unscheduled maintenance for the seals, bearings and impeller will be quite high, up to 70% of the cost of a replacement in some applications. These costs do not take in account the extra costs of removing the pump from line and disruption of the process (Lou et al., 2012). Therefore, it is worth to increase the circulation rate to save the expensive pump, to lower maintenance costs and to minimise the cost of process disruptions.

The temperature of the absorption column is controlled by adjusting the lean amine temperature entering the column from the top. Lean amine temperature can be controlled by varying the speed of the cooler fans. The current lean amine temperature is around 57°C. A sensitivity analysis was carried out using HYSYS (Fig. B4) to investigate the effect of lean amine temperature on the performance of the acid gases removal, keeping the solvent circulation flow rate (18.75 m^3 hr^{-1}), amine concentration (41 wt. %), inlet feed gas conditions and other parameters constant. Increasing lean amine temperature will increase the CO_2 removal efficiency, while H_2S removal efficiency will decrease. There is a sort of competing effects between the solubility and reaction kinetics of CO_2 with MDEA. Removal efficiency of CO_2 reaches its maximum at temperatures between 60-65°C. The kinetic effects in this range tend to dominate. Above 65°C the solubility limitation starts to take place. On the other hand, MDEA loses its ability of absorbing H_2S when the column temperature rises. Therefore, as H_2S is the main acid gas that needs to be removed from feed gas, low column temperature is favoured. However, lowering the absorber temperature (Fig. B5) increases the absorption of hydrocarbons as well as the regeneration reboiler requirement. Moreover, it was found that increasing the absorption column temperature will increase the losses of water and MDEA to the sweet gas (Fig. B6). There is a water loss of approximately 0.5 kmol h^{-1} for every 5°C increase in temperature. The sweet gas exits from the absorber as saturated with water at high column temperatures. In this case, the role of thumb evaluated by Addington and Ness (2013), which says that the lean amine entering the absorber has to be 5°C warmer than the feed gas, needs to be applied. This is to avoid any condensation of light hydrocarbons constituents of the feed gas and to ensure that no phase change occurs. Therefore, as the feed gas enters at 51°C the lean amine temperature will be fixed at 56°C.

The plant was designed for an MDEA concentration of 38.5 wt. % (PDO Company, 2012), but after the start up and troubleshooting it has been fixed at the current MDEA concentration of 41 wt. %. The concentration is controlled by adjusting the make-up water. The CO_2 and H_2S concentrations in the sweet gas for different MDEA concentrations showed that H_2S concentration remained relatively unchanged with increasing MDEA concentration. On the other hand, CO_2 concentration decreased by around 1.0 ppm with each 5 wt. % increase in MDEA concentration. More CO_2 absorbed can displace H_2S. Nonetheless, the H_2S concentration is always within the pipeline specifications and still below the maximum allowed concentration (5 ppm). However, it seems that HYSYS over-estimates the real concentration of H_2S in the sweet gas. The only way to prove this is through using the available sensors and laboratory tests by taking samples after each change. After this investigation, the MDEA concentration should be

returned to its design specification at 38.5 wt. %. Table 5 summarizes the optimized operating parameters.

Table 5

Optimized operating parameters for the gas sweetening plant.

Parameter	Current plant data	Optimized value
Circulation rate (m^3 hr^{-1})	18.75	40.0
Lean solvent temperature (°C)	57.0	56.0
MDEA concentration (wt. %)	41.0	38.5

3.3. Process modifications

The conventional split-loop process modification (Case 2) was simulated using Aspen HYSYS as shown in Figure 2. Following the optimization algorithm shown in Figure 3, the side fraction was increased starting from 0.1. The maximum fraction that could be achieved was 0.7. Below this fraction, the MDEA concentration in the lean amine stream was below 41 wt. %, and hence it could not be controlled by make-up water. The industrial practice is to evaporate some water to increase MDEA concentration and hence a reboiler is required. Allowing the removal of more fluid in the side draw (Fig. B7) leads to a considerable decrease in the regeneration duties (reboiler and condenser). However, more H_2S had slipped to the sweet gas stream, which was due to unstripped acid gases in the side draw. Therefore, 0.7 side fraction was selected. At this fraction, there were big savings of around 1480 kW in the stripper heating requirement and around 1450 kW in the cooling requirement. However, the H_2S concentration in the sweet gas is around 1.0 ppm, which is still less than the upper limit of 5 ppm.

The next step in the algorithm (Fig. 3) was to examine the effect of side-draw stage location in the stripper and the side-feed stage location in the absorber on the absorption of H_2S. Their selections are optimized in order to give lower H_2S concentration in the sweet gas. The examination showed that drawing fluid from the upper stages of the stripper gave lower H_2S concentration in the sweet gas. H_2S concentration in the sweet gas stream from the absorption column increased as the side-draw stage location was approaching the inlet stage of the feed to the stripper. As the difference was relatively minor, the middle stage (the 9^{th} from the bottom) was selected. In the absorber column, the same relationship was figured out until reaching stage number 15, where H_2S concentration starts to significantly increase. The location of the side feed is preferred to be in the middle and hence it was again located at the middle stage (the 13^{th} from the bottom) of the absorber.

The last step in the algorithm was adjusting the reboiler duty to meet the H_2S specification in the sweet gas. As the H_2S concentration was 1.0 ppm, which is well below the maximum allowable concentration, the reboiler duty would not be changed.

There are some constraints or limitations that need to be considered in the conventional split-loop process modification. The simulation had shown the possibility of introducing vapour to the side draw pump, which can cause entrainment (Ibrahim et al., 2014). The side draw from intermediate stages is a mixture of liquid and vapour and if the vapour fraction was high, a separator or a condenser has to be installed to remove vapour. The minimum temperature

approach in the heat exchanger (lean/rich amine streams) is 10°C. So, a simultaneous and careful monitoring including pinch analysis is required.

The Aspen HYSYS simulation of the modified split-loop process (Case 3) is shown in Figure 4. In this modified process, the side draw from the stripper was around 34 wt. % MDEA concentration, so it had to be increased to about 38.5 wt. %. Therefore, it was firstly heated where a portion of water would be vaporized in order to adjust the concentration of MDEA to the desired value. After trial and error, the temperature of the proposed heater should be around 123.4°C. The water vapour outlet of the flash separator was returned to the stripper. This flash vessel can also decrease the acid gases in the side draw. It was found that, at any side draw flow rate, around 40 mol % of H_2S and more than 60 mol % of CO_2 would leave with the water vapour.

To apply the optimization algorithm (Fig. 3), the side fraction was initially increased and the H_2S concentration in the sweet gas and regeneration duties were recorded (Fig. B8). The results showed that H_2S concentration increased with increasing the side draw flow rate due to the increased unstripped acid gases. The heating requirement in the stripper decreased considerably. However, the cooling requirement increased with increasing side draw fraction. This could be due to the water vapour returned to the column. The selected stage location of the water vapour return was between the side-draw stage (the 9^{th} from the bottom) and the stripper feed stage (the 17^{th} from the bottom up). However, changing the location of the water vapour return did not make a big saving or change in the sweet gas purity. Therefore, its location was kept at the middle stage (the 13^{th} from the bottom). Again, the maximum side fraction to be achieved was 0.7. To select the optimal side fraction, two fractions were investigated in terms of energy consumption and H_2S concentration. At these two fractions, H_2S concentration was well below the allowable maximum concentration. At a side fraction of 0.2, the heating and cooling requirements of the stripper were equal at 1930 kW. At 0.7, reboiler duty decreased by 1110 kW, while the condenser increased by only around 578 kW and this increase was gradual. Taking into account the higher operating cost of heating, a side fraction of 0.7 was selected.

For the purpose of heat integration, the side draw outlet temperature of the side draw pump, which is 125.8°C, needs to be cooled to around 55°C (the temperature of the side-draw location). A pinch analysis is required to set the process heat recovery and to avoid any heat transfer across the pinch by inappropriate use of utilities. The heat can be added to the cold rich amine stream via a process-to-process heat exchanger as shown in Figure 4 (E-102 equipment). As described by Smith (2005), the heat exchanger network design for maximum energy recovery can be represented by a grid diagram (pinch analysis). Figure 6 is a grid diagram that shows the maximum recovery between the hot side-draw stream and the cold rich amine through a heat exchanger at a minimum approach temperature (ΔT_{min}) of 10°C. The network is divided into above and below the pinch. Using Problem Table (Smith, 2005) method and SPRINT software, the pinch temperature was found to be 56.25°C. To satisfy the heating and cooling requirements, one process-to-process heat exchanger is required with a maximum heat recovery of around 864.97 kW, minimum cold utility of 83.75 kW and minimum hot utility of 42.88 kW.

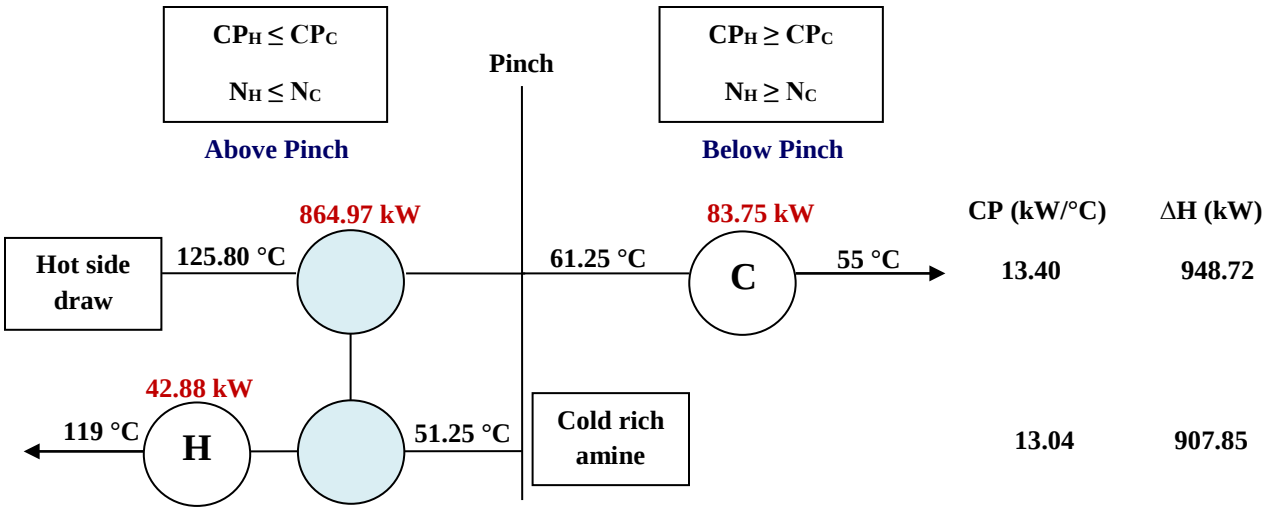

Fig. 6. Pinch analysis for rich/side draw amine heat exchanger (E-102 equipment in Fig. 4).

For the last step in the algorithm, the H_2S concentration was 0.41 ppm, which is well below the maximum allowable concentration in the sweet gas. Therefore, no need to adjust the reboiler duty.

Table 6 gives an overall comparison between the three cases with a summary of their cost estimations. It is apparent from this table that Case 2 was the least harmful process to the environment with a carbon footprint of 424.93 kg CO_2e per 1.0 ppm H_2S compared to more than 1000 kg CO_2e per 1.0 ppm H_2S in the other two processes. This process could be considered as the most sustainable process among the three processes. Even though the process modifications are mainly retrofit-based, the capital costs were included since these modifications require investing into some additional units and equipment. As can be seen from Table 6, Case 2 has annual operating costs of around £0.116 MM (50% saving compared to the base case) with an increase of annualized capital cost by only 8.4%. Overall, there is a total annualized saving of around £0.06 MM yr^{-1}. The base plant (Case 1) already exists and the equipment have been already purchased. Therefore, in terms of payback period of this modification, and assuming the energy saving as positive, the payback period is about 1.5 years only. Therefore, in addition to its low gas emissions, the split-loop process modification (Case 2) for the current plant is beneficial and recommended. On the other hand, Case 3 requires higher investment, and thus it is not the optimal choice.

Table 6

Comparison among the base case (Case 1), conventional split-loop (Case 2) and modified split loop (Case 3).

Parameters	Case 1 (Fig. 1)	Case 2 (Fig. 3)	Case 3 (Fig. 4)
Side fraction	0.0	0.7	0.7
Stripper intermediate side draw stage [a]	--	9	9
Absorber intermediate side feed stage [a]	--	13	13
H_2S concentration in sweet gas (ppm)	0.24	1.00	0.41
Utilities (kW)			
Total Heating requirement	2170	690	2733[b]
Total Cooling requirement	1961	513	2508
Total Power requirement	147	176	136
Carbon footprint (kg CO_2e/1 ppm H_2S)			
MP steam [c]	0.00	0.00	616.34
Hot oil [d]	803.35	255.82	319.11
Cooling process [e]	511.33	133.77	653.95
Electricity [f]	99.98	119.71	92.50
Total	1414.66	509.30	1681.90
Profitability analysis			
ISBL (£MM*)		0.121	0.359
OSBL (£MM)		0.036	0.108
Engineering (£MM)		0.008	0.023
Contingency (£MM)		0.031	0.093
Total Capital Fixed costs (£MM)		0.196	0.583
Working Cost (£MM)		0.029	0.087
Total Capital Investment (£MM)		0.225	0.670
Total Capital (September 2014) (£MM)		0.273	0.813
Annualized capital cost (£MM yr^{-1})		0.091	0.271
Total Operating Cost (£MM yr^{-1})	0.230	0.116	0.199
Total Annualized cost (£MM yr^{-1})	0.230	0.207	0.470

[a] Stages numbered from bottom up

[b] Most contribution to this duty is from the heater (E-101) in Fig 4 (around 1870 kW).

[c] Steam emits 0.361 kg CO_2e per kWh energy (PDO figures)

[d] Hot oil emits 0.405 kg CO_2e per kWh energy (PDO figures)

[e] Process cooling emits 0.292 kg CO_2e per kWh energy (PDO figures).

[f] Electricity emits 0.746 kg CO_2e per kWh energy (PDO figures)

* MM stands for Million

3.4. Sulphur recovery process

Figure 5 shows the Aspen HYSYS simulation of a conventional sulphur recovery process (Case 4). The acid gas stream from the stripper and the stoichiometric amount of oxygen required to burn almost one-third of the hydrogen sulphide to sulphur dioxide are fed through a furnace, where at 1150°C, a substantial amount of elemental sulphur is formed, with typically 60 to 70% of the H_2S in the feed gas converted to sulphur. This sulphur is condensed, separated and routed to the sulphur collector (Mix 100). The gases leaving the sulphur condenser are reheated and fed to the first catalytic reactor, where more sulphur is produced. The Catalyst (DD-431) is the common catalyst used in sulphur recovery process (Asadi et al., 2013). The gases leaving Reactor1 are again cooled and sulphur is condensed and routed to the sulphur collector. After the sulphur removal, the gases are reheated again before entering the second catalytic reactor (Reactor2). The outlet gases from the second reactor are again cooled and more sulphur is condensed and routed to the sulphur collector. The exhaust outlet gases from Splitter3, which is almost free of H_2S, are sent to incinerator before being flared. With an almost total conversion of H_2S, the total sulphur produced in the process is around 336 kg day^{-1}.

The process can be simulated here but in reality there could be unstable operation in the furnace due to the low acid gases in the feed which can be incapable of producing stable flame in the furnace (Mahdipoor et al., 2012). Therefore, the inlet stream has to be modified to ensure the process stability. The best modifications are either enriching the inlet air with oxygen or bypassing some of the feed acid gas around the furnace, or combining both options. Figure 7 shows the bypass modification of the conventional sulphur recovery process. After trial and error with these options, the two selected processes were; 50% acid gas bypassed with 30% O_2 in the air (named as Case 4-2) or 35% acid gas bypassed with 100% O_2 in the air (named as Case 4-3). The outlet temperatures of the furnace in Case 4, Case 4-2 and Case 4-3 are 1149°C, 1018°C and 1004°C, respectively.

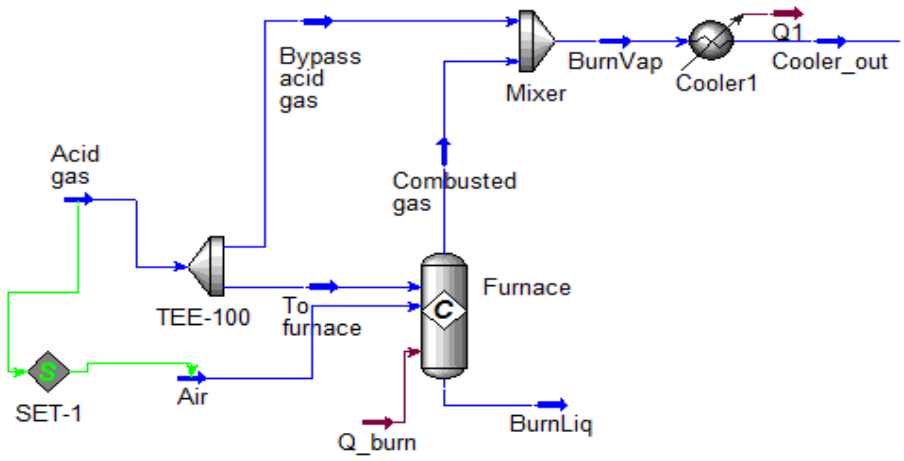

Fig. 7. Aspen HYSYS simulation of the bypass modification of the conventional sulphur recovery process.

As the sulphur recovery process is highly exothermic, valuable steam can be generated. The steam generated in the sulphur recovery conventional process is a medium-pressure (MP) steam for Cooler1 and a low-pressure (LP) for Cooler 2 and Cooler3. Looking at the outlet hot temperatures required in the two heaters (Heater1 and Heater2) in Figure 5 (260°C and 210°C,

respectively), these steams cannot be utilized for heat integration due to their low saturation temperatures and they cannot also be utilized in the modified split-loop process (Fig. 4). Total Site Profile method can be used to investigate and target steam generated to be used by different processes (Smith, 2005). STAR software was used assuming a minimum approach temperature of 10°C. The Total Site Profile method showed that there is no site sink profile (positive enthalpies). Therefore, the generated steam cannot be used as a hot utility in these processes.

Table 7 summarizes the profitability analysis of the conventional sulphur recovery process (Case 4) and the two selected modifications (Case 4-1 and Case 4-2). As can be seen from the table, the values of NPV are negative and the payback periods are too long, due to the negative cash flows and low price of sulphur in markets. Therefore, in terms of profitability, the proposed sulphur recovery process is basically not beneficial and would be rejected. However, as mentioned earlier, the main function of this proposal is to reduce the emissions of the waste gases and subsequently making the gas sweetening process more sustainable. So, the sulphur recovery unit is needed. Overall comparison of the three cases, the bypassed stream with entrenchment of 100% O_2 (Case 4-2) can be considered as the best process. In terms of the total annualized costs, Case 4-3 is the cheapest. In terms of sulphur production, it also produces sulphur 46% and 26% higher than Case 4 and Case 4-2, respectively. Moreover, this process is more sustainable as its carbon footprint in the production stage is the lowest. This is consistent with the findings of Mokhatab et al. (2008). However, higher oxygen enrichment level requires special oxygen piping and burner technology (Manenti et al., 2014), which would require some additional investment. This has not been considered in this study.

Table 7

Comparison between the conventional sulphur recover process and its two modifications.

	Conventional process (Case 4)	50% acid gas bypassed with 30% O_2 (Case 4-1)	35% acid gas bypassed with 100% O_2 (Case 4-2)
Total Sulphur produced (kg day^{-1})	336.0	291.0	426.0
Waste gas flared (kg h^{-1})	874.0	882.8	929.5
Utilities (kW)			
Total cooling duty	327.81	232.86	252.86
Total heating duty	39.21	39.44	40.12
Carbon footprint (kg CO_2e/kg sulphur)			
MP steam	0.93	1.08	0.75
Cooling process	6.12	5.02	3.71
Waste gas flared *	136.20	158.83	114.19
Total	143.25	164.93	118.65
Profitability Estimation			
ISBL (£MM)	26.815	21.146	22.145
OSBL (£MM)	8.044	6.344	6.644
Engineering (£MM)	1.743	1.375	1.439
Contingency (£MM)	6.972	5.498	5.758
Total Capital Fixed costs (£MM)	43.574	34.363	35.986
Working Cost (£MM)	6.536	5.154	5.398
Total Capital Investment (£MM)	50.110	39.517	41.384
Total Capital (September 2014) (£MM)	60.748	47.906	50.169
Annualized capital cost (£MM yr^{-1})	20.249	15.969	16.723
Total Operating Cost (£MM yr^{-1})($\times 10^{-3}$)	0.005	0.005	0.005
Total Annualized cost (£MM yr^{-1})	20.254	15.974	16.728
NPV	-336.0	-291.0	-426.0

* Hazardous waste incineration emits 2.39 kg CO_2e per kg waste (PDO figures).

4. Conclusions

The existing ammonia (MDEA) gas sweetening plant in Lekhwair (Oman) can be modelled and optimized using commercial process simulator Aspen HYSYS v7.3. Two fluid packages were investigated firstly and the improved DBR amine package introduced to Aspen HYSYS v7.3 showed to more suitable for the simulation and sensitivity analysis of MDEA gas sweetening process. The simulation results using DBR amine package were almost matching the actual plant data, with absolute error less than 5% for sweet gas stream specifications and less than 2% for required utilities. After reviewing the current operation of the existing plant and considering the

main operating parameters (lean solvent circulation flow rate, lean amine temperature, and MDEA concentration), it was found that the circulation rate needs to be increased to at least 40 $m^3\ h^{-1}$ to avoid damaging the pump, lean amine temperature to be kept unchanged at 56°C to avoid foaming in the system, and the MDEA concentration to be decreased to the design specification of 38.5 wt. %. To save some operating costs in the system, taking the advantage of the 5 ppm H_2S concentration allowance in the sweet gas, two modifications (conventional split-loop and modified split-loop with heat integration) were proposed. An optimization procedure was followed to achieve the maximum savings and to avoid equipment limitations. It was found that the conventional split-loop modification can save up to 50% of the current operating expenses with only around £175,000 increase in the capital investment, and a penalty of 1.0 ppm of H_2S concentration in the sweet gas, which is still well below the pipeline gas specification. Around 58800 kmol yr^{-1} of waste gases from the gas sweetening process are burnt and flared to the atmosphere. Therefore, a sulphur recovery unit was simulated and optimized by proposing process enhancements based on gas-bypass configuration. It was found that a sulphur recovery unit with 35% bypassed gas stream over the furnace and 100% O_2 enrichment in the air stream can reduce the gas emissions and produce valuable sulphur.

Additional studies might be performed to further optimize the plant. Such studies might include examining the effect of using different amines, or investigating the pre-treatment of the sour gas and the sensitivity analysis of its pressure, temperature, flowrate, and other parameters taking into consideration the physical and chemical properties of the components that may influence the efficiency of the gas sweetening process.

Acknowledgment

The authors would like to acknowledge Centre for Process Integration at the University of Manchester for their technical support, especially providing an access to their own optimization software (SPRINT v2.7 and STAR v2.7, 2011).

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Appendix A: Cost estimation (Sinnott and Tawler, 2009)

A1. Capital costs

The total capital investment is the sum of the fixed capital investment and the working capital. The fixed capital investment of a chemical process plant includes:

- The Inside Battery Limits (ISBL) which is the purchase cost of the major equipment items required for the process, and the other installation costs being estimated as factors of the equipment cost.
- The Outside Battery Limit (OSBL) cost which includes the cost of additions to the site infrastructure or increasing the capacity of an existing plant. It will be assumed as 30% of ISBL.
- Engineering and contingency expenses. For the preliminary design, where only the main equipment are considered, contingency and engineering charge of 25% of (ISBL+OSBL) cost will be assumed on this project.

The fixed capital investment is the sum of ISBL, OSBL, engineering and contingency. The working capital is the capital required to actually operate the plant. A charge of 15% of the total capital investment will be considered for the working capital cost. Start-up cost of 10% of the fixed capital cost will be included in the costing of the sulphur recovery unit. The accuracy of this type of estimate depends on the stage of the design and on the reliability of the data available on equipment costs. There are many methods used for cost estimate. The factorial method is a relatively accurate method for capital cost estimation and is followed in this project.

The preliminary purchased cost of an equipment can be estimated using the following correlation:

$$Ce = a + bS^n \qquad \text{(A.1)}$$

Where C_e = purchased equipment cost; S = size parameter (varied units); a, b = cost constants shown in Table A1; and n = exponent for that type of equipment shown in Table A.1.

Table A.1

Purchased Equipment cost for the main equipment in gas sweetening process

Equipment	Units for Size, S	a	b	n
U-tube shell and tube	Area, m^2	10,000	88	1.0
Kettle reboiler	Area, m^2	14,000	83	1.0
Centrifugal pump	Flow, L/s	3,300	48	1.2
Electric driver	Power, kW	920	60	0.7
Vertical Pressure vessel, CS	Shell mass, kg	-400	230	0.6
Trays (Bubble cap)	Diameter, m	200	240	2.0
Trays (Sieve)	Diameter, m	100	120	2.0
Reactor	Volume, m^3	14000	15,400	0.7

The area of a heat exchanger can be obtained from the overall heat balance equation as:

$$A = \frac{Q}{U\Delta T_{lm}} \tag{A.2}$$

Where Q = the heat duty of the heat exchanger (kW), U = overall heat transfer coefficient (kW m^{-2} °C^{-1}), and ΔT_{lm} = log-mean temperature difference (°C).

As can be seen from Table A.1, there is no correlation for the distillation and absorption columns. The column is a combination of a vertical pressure vessel and internals (trays). The vessel wall thickness and, subsequently, the shell mass can be calculated using the guide explained by Sinnott and Tawler (2009).

Note that the estimated costs of purchased equipment are based on Chemical Engineering (CE) index (January 2006). To get the present costs, the following cost escalation method (Eq. A.3) is used to relate the present costs to past costs. The CE index for January 2006 is 478.6 and for preliminary September 2014 is 580.2.

$$Cost\ in\ year\ A = Cost\ in\ year\ B\ x\ \frac{Cost\ index\ in\ year\ A}{Cost\ index\ in\ year\ B} \tag{A.3}$$

The installation factors include equipment erection, piping, instrumentation, electrical, civil and paint. The installation factors for the main equipment used in this project are given in Table A.2.

Table A.2

Installation factors

Equipment	Installation factor
Distillation column	4
Heat exchangers	3.5
Pressure vessels	4
Pumps	4
Reactors	4

However, these factors are for plants built from carbon steel. As some equipment in this project will be constructed using stainless steel, as shown in Table A.3, the capital cost of each equipment has to be multiplied by 1.3.

Table A.3

Materials of construction of the main equipment in the gas sweetening process

	Absorber (Shell & Trays)	Stripper (Shell & Trays)	Reboiler	Condenser	Lean/Rich HEX	Air cooler	Booster & Charge lean amine pumps
Material of Construction	CS	SS	SS	SS	SS	CS	CS

CS = Carbon Steel, SS= Stainless Steel

It is a common practice to report operating and capital costs on annualized basis. The capital cost can be annualized by introducing a capital charge factor (CCF), where the CCF includes all capital-related expenses (repair, maintenance, etc.). For preliminary design, CCF of 1/3 yr is considered.

A2. Operating costs

The operating costs include mainly utilities (fuels, electricity, steam, and cooling water) which are the most significant variable operating costs, catalysts for sulphur recovery process, and chemicals (MDEA amine and antifoam chemical). Steam is considered available at different pressure levels (HP, MP, and LP). The steam costs are derived using a simple enthalpy method of steam costing.

Cooling water is assumed to be 1.0% of the cost of power, and the cost of electricity is considered as \$0.045 kWh^{-1}. For a year with 8600 hours; the cooling water cost is calculated as $0.045 \times 0.01 \times 8600 = \$3.87\ kW^{-1}\ yr^{-1}$.

Hot oil is currently used in stripper reboiler. The cost of hot oil can be taken as the cost of providing heat to the fuel in the fired heater. The fuel considered is natural gas at a price of \$4.9 per thousand cubic feet, which can be converted to \$140.27 $kW^{-1}\ yr^{-1}$. Table A.4 lists the average costs of utilities in Great Britain's Pound Sterling currency, where 1 GBP equals 1.54 USD (Average exchange rate of 2014).

Table A.4

Costs of utilities used in the process

Utility	Cost
MP steam	£70 $kW^{-1}\ yr^{-1}$
LP steam	£52 $kW^{-1}\ yr^{-1}$
Electricity	£251.3 $kW^{-1}\ yr^{-1}$
Cooling water	£2.51 $kW^{-1}\ yr^{-1}$
Hot oil	£91.05 $kW^{-1}\ yr^{-1}$
Fuel	£0.046 kg^{-1}
MDEA	£0.286 kg^{-1}

Appendix B: Aspen HYSYS Sensitivity analysis

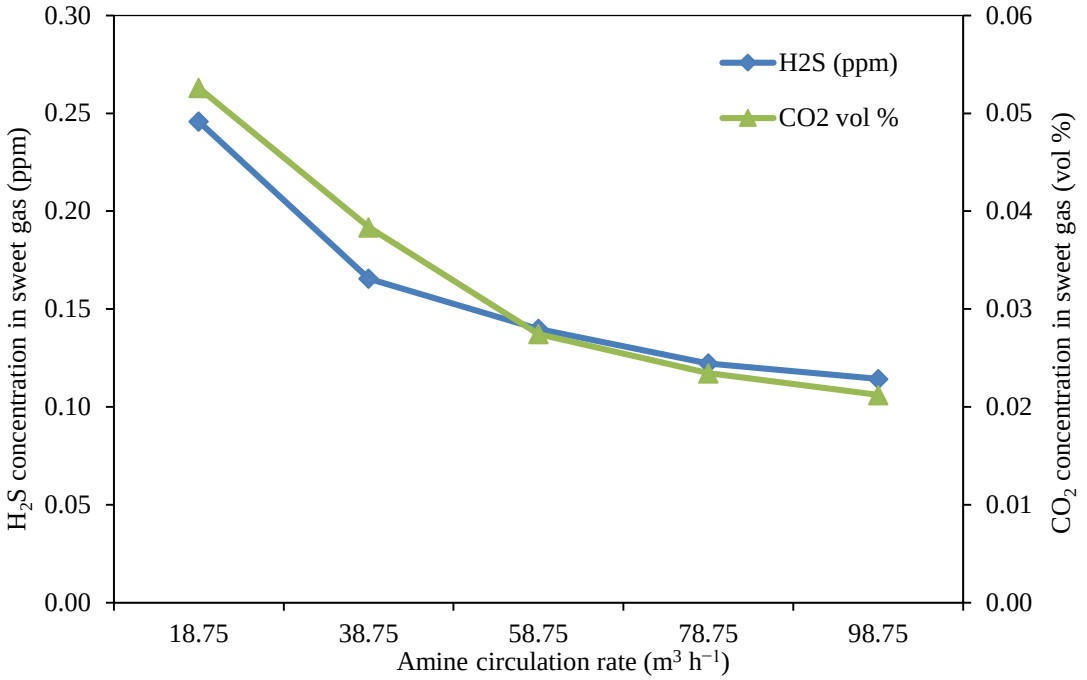

Fig. B.1. Effect of MDEA circulation rate on acid gases concentration in sweet gas.

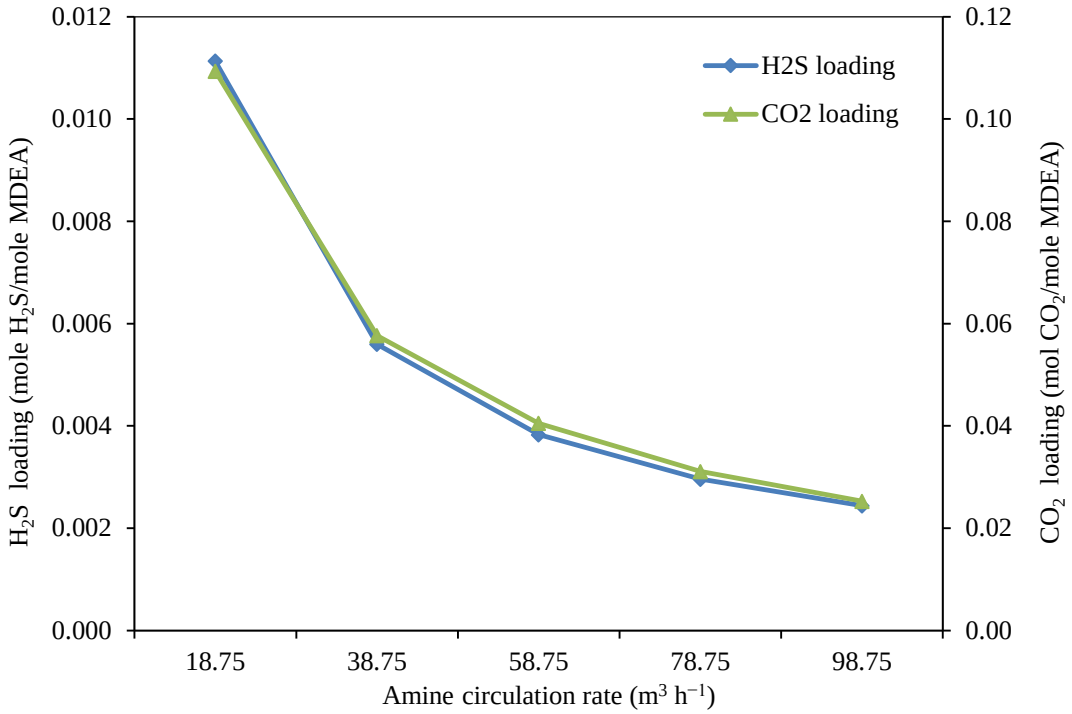

Fig. B2. Sensitivity analysis of the effect of amine circulation rates on acid gas loading in the rich amine stream.

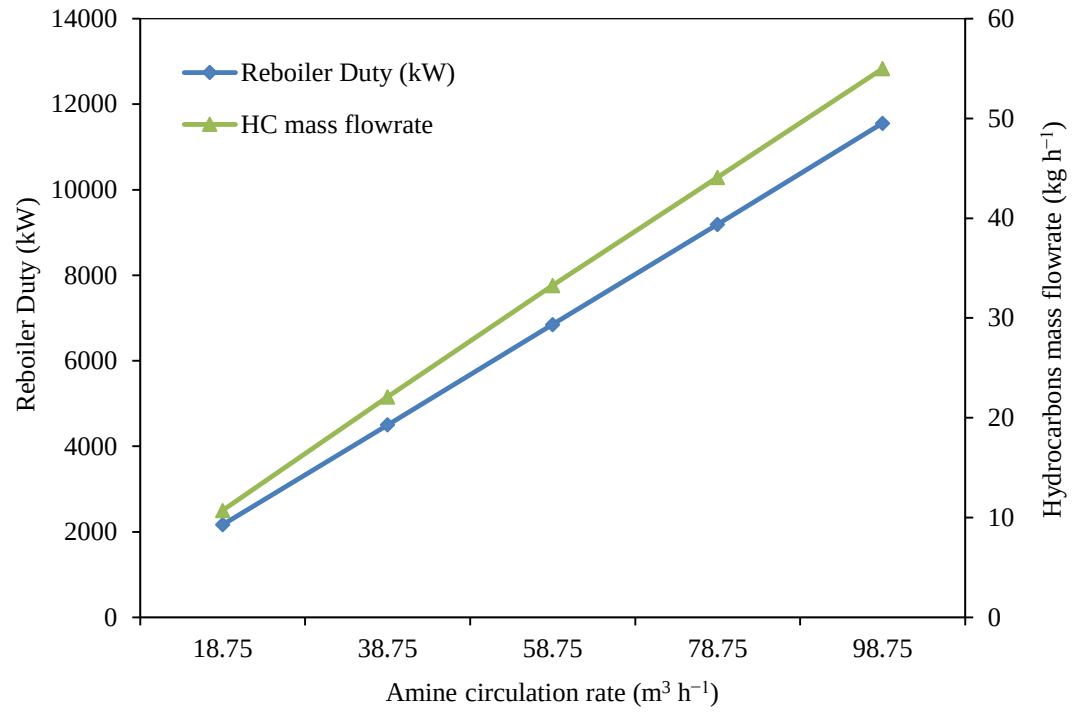

Fig. B.3. Sensitivity analysis of the effect of amine circulation rates on the reboiler duty and hydrocarbons mass flowrate in the rich amine stream.

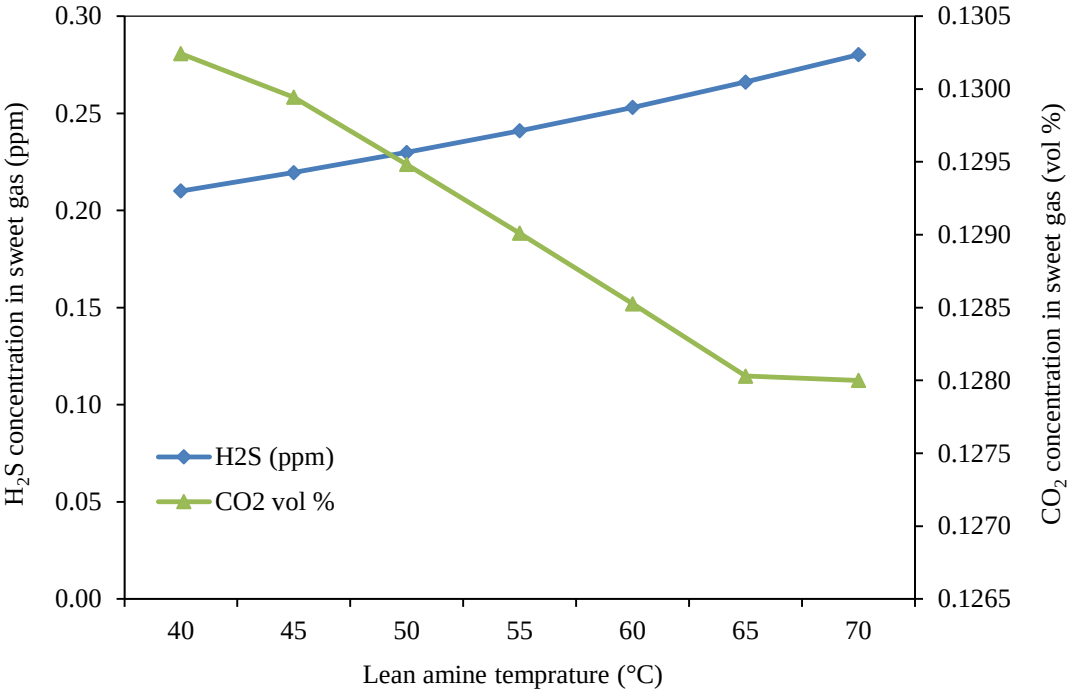

Fig. B.4. Sensitivity analysis of the effect of lean amine temperature on acid gases concentration in sweet gas.

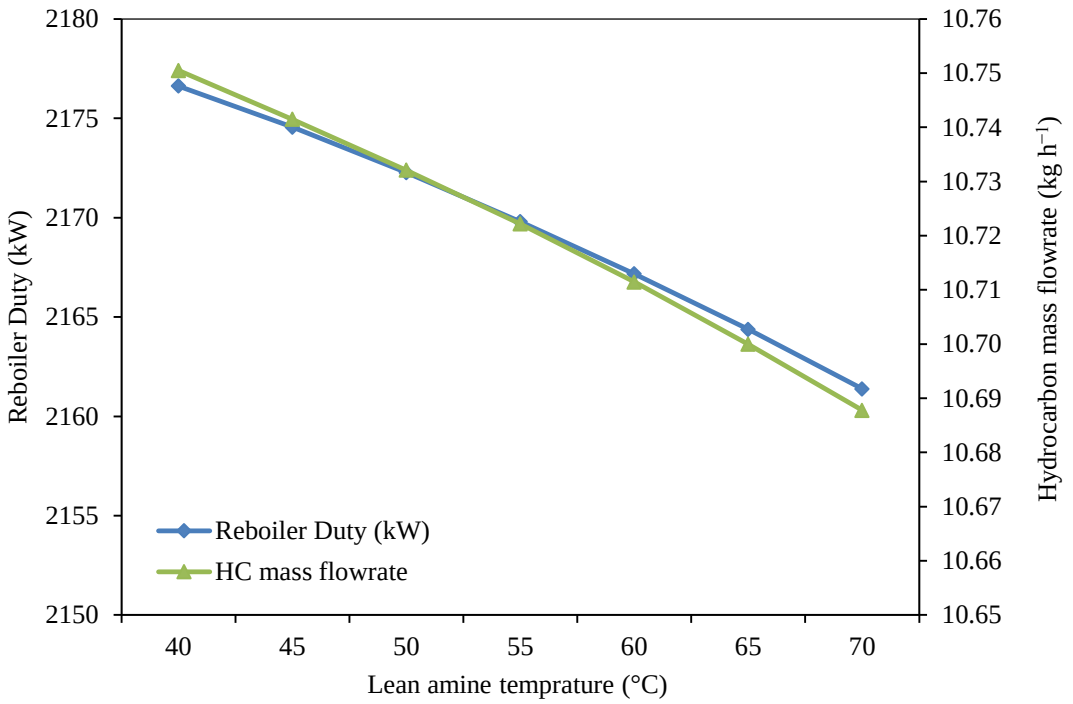

Fig. B.5. Sensitivity analysis of the effect of lean amine temperature on the reboiler duty and hydrocarbons mass flowrate.

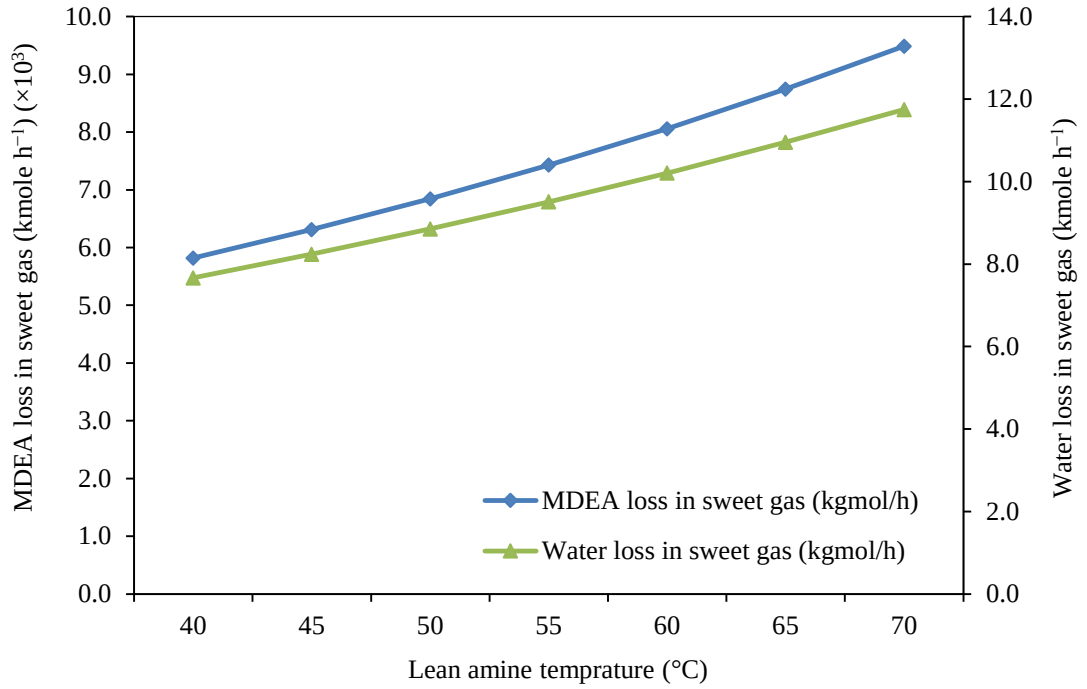

Fig. B.6. Effect of lean amine temperature on the losses of MDEA and water in the sweet gas stream.

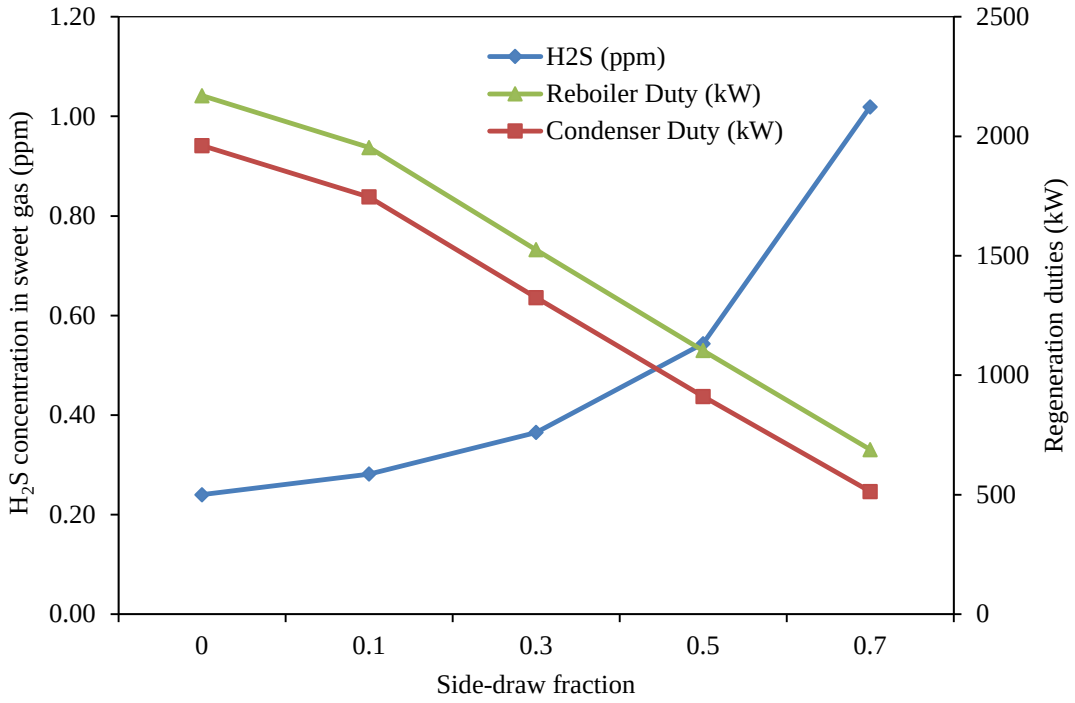

Fig. B.7. Effect of side-draw fraction on the H_2S concentration in the sweet gas and regeneration duties (conventional split-loop).

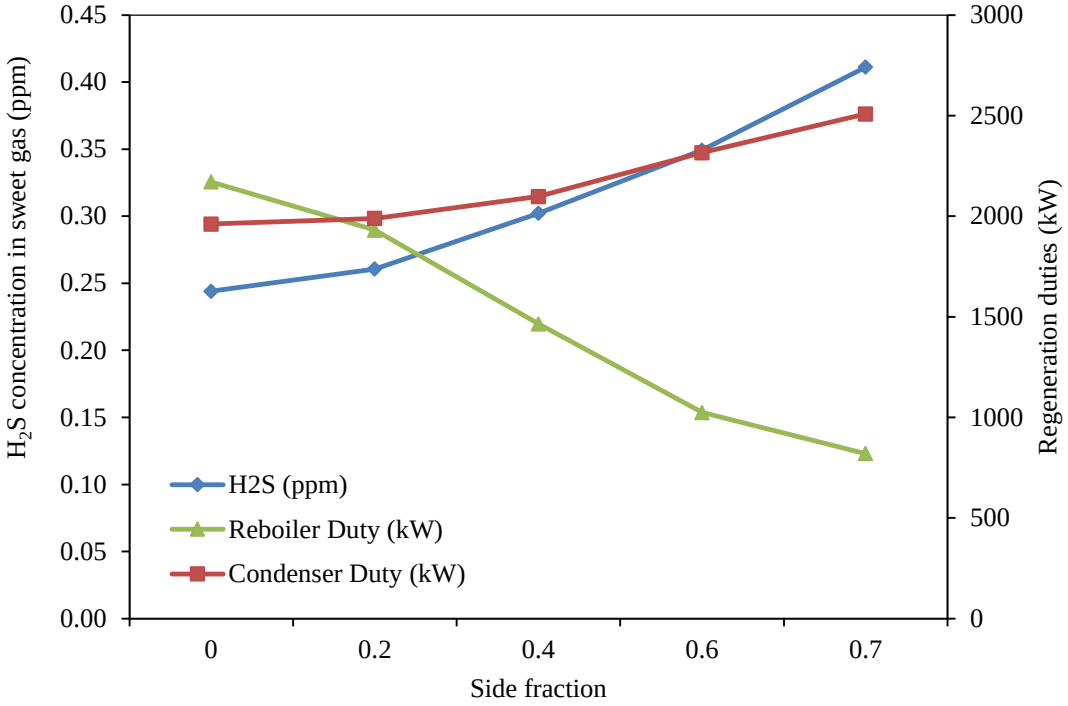

Fig. B.8. Effect of side-draw fraction on the H_2S concentration in the sweet gas and regeneration duties (modified split-loop).